PROCESS AND APPARATUS FOR CONTINUOUS POLYCONDENSATION

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TECHNICAL FIELD

5   The present invention relates to a process and an apparatus for continuous polycondensation of polymers of polyester series such as polyethylene terephthalate, polybutylene terephthalate, etc., from an aromatic dicarboxylic acid or its derivative and a glycol.

BACKGROUND ART

    According to the conventional process for producing polymers of polycondensation series of polyethylene terephthalate, etc., raw materials terephthalic acid and ethylene glycol are fed to a mixing tank in an appropriate proportion for esterification and

15   the mixture is then fed to esterification reactor tanks through a pump. In the esterification step, 2 or 3 stirring tanks each provided with stirring vanes are provided in series and a water effluent as a by-product is separated in a distillation column. A

20   prepolymerization step follows, where a plurality of vertical stirring tanks or horizontal stirring tanks are provided, and a final polymerization step follows, where a horizontal stirring tank is provided. The tanks of these polymerization steps are each provided with a

condeser for removing an effinent ethylene glycol as a
by-product, and are operated in a reduced pressure
atmosphere.   In the conventional process for producing
polyester there are 4 to 6 reactor tanks, each of which
5   is provided with strring vanes and its power source and
further provided with a distillation column or a
condenser for separating and removing the by-product.
Furthermore, the polymerization steps are operated in a
reduced pressure atmosphere, and thus the necessary
10   additional vacuum means for generation of a reduced
pressure is required.   That is, the operation of the
apparatus requires a higher maintenance cost and a higher
apparatus cost.   One example of the related art is
disclosed in JP-A 7-207009, where each tank is provided
15   with stirring vanes and is vacuum-controlled, and thus
its running or operation is complicated.   That is,
improved apparatus and process with a higher efficiency
throughout the apparatus and economical operation based
on energy saving throughout the apparatus have been so
20   far desired for the production of high molecular weight
polyester.

DISCLOSURE OF THE INVENTION

       An object of the present invention is to
provide a process and an apparatus for continuous
25   polycondensation, capable of efficiently producing a high
molecular weight polyester.

       Another object of the present invention is to

provide a process and an apparatus for continuous
polycondensation, capable of efficiently producing a high
molecular weight polyester with simple apparatus
structure and arrangement.

5   Other object of the present invention is to
provide an apparatus for continuous polycondensation and
a process for continuous polycondensation, which can
overcome the problems of the above-mentioned prior art
and can conduct more efficient reaction to produce
10 polymers of good quality with a minimum energy through
minimum necessary reactor structure and arrangement.

          According to a first aspect of the present
invention, a process for continuously producing polyester
is provided, which comprises a first step of reacting an
15 aromatic dicarboxylic acid or its derivative with a
glycol in a first reactor, thereby producing an
oligoester or a polyester having an average degree of
polymerization of 3 to 7; a second step of polycondensing
the oligoester or polyester of the first step in a second
20 reactor, thereby producing a low molecular weight
polyester having an average degree of polymerization of
20 to 40; and a third step of further polycondensing the
low molecular weight polyester of the second step to an
average degree of polymerization of 90 to 180 in a third
25 reactor, thereby producing a high molecular weight
polyester, wherein at least one of the first reactor and
the second reactor is free from an external power soure
stirring means; or wherein a reactor is used as the third

reactor, which comprises a horizontal cylindrical vessel having an inlet for the low molecular weight polyester of the second step at the lower part at one end thereof and an outlet for the high molecular weight polyester at the lower part at the other end thereof in the longitudinal direction thereof, an outlet for volatile matters at the upper part thereof, and a stirring rotor rotatable in the vicinity of the inner periphery of the vessel in the longitudinal direction of and within the vessel, the stirring rotor within the vessel comprising a plurality of stirring blocks according to the viscosity level of the polyester, and the stirring rotor having stirring vanes without any rotating shaft at the center of the stirring rotor; more preferably wherein a mixture of the aromatic dicarboxylic acid or its derivative and the glycol in a molar ratio of 1:1.05 to 1:2.0 of the former to the latter is fed to the first reactor kept at a temperature of 240° to 285°C under a pressure of the atmospheric pressure to 3×10^5 Pa; the oligoester or polyester from the first reactor is fed to the second reactor kept at a temperature of 250° to 290°C under a pressure of the atmospheric pressure to 133 Pa; and the low molecular weight polyester from the second reactor is fed to the third reactor kept at a temperature of 270° to 290°C under a pressure of 200 to 13.3 pa; more preferably wherein the stirring rotor is rotated at 0.5 to 10 rpm; and more preferably wherein total reaction time throughout the first reactor, the second reactor and the

third reactor is 4 to 8 hours.

According to a second aspect of the present invention, an apparatus for continuously producing polyester is provided, which comprises a first reactor

5 for reacting an aromatic dicarboxylic acid or its derivative with a glycol, thereby producing an oligoester or polyester having an average degree of polymerization of 3 to 7; a second reactor for polycondensing the oligoester or polyester from the first reactor, thereby

10 producing a low molecular weight polyester having an average degree of polymerization of 20 to 40; and a thrid reactor for further polycondensing the low molecular weight polyester from the second reactor to an average degree of polymerization of 90 to 180, thereby producing

15 a high molecular weight polyester, wherein at least one of the first reactor and the second reactor is free from an external power source stirring means; or wherein the third reactor comprises a horizontal cylindrical vessel having an inlet for the low molecular weight polyester

20 from the second reactor at the lower part at one end thereof and an outlet for the high molecular weight polyester at the lower part at the other end thereof in the longitudinal direction thereof, an outlet for volatile matters at the upper part thereof, and a

25 stirring rotor rotatable in the vicinity of the inner periphery of the vessel in the longitudinal direction of and within the vessel, the stirring rotor within the vessel comprising a plurality of stirring blocks

according to the viscosity level of polyester feed and the stirring rotor having stirring vanes without any rotating shaft at the center of the stirring rotor; more preferably wherein the first reactor is conduct the

5 reaction at a temperature of 240° to 285°C under a pressure of the atmospheric pressure to 3×10^6 Pa; the second reactor is to conduct the polycondensation at a temperature of 250° to 290°C under a pressure of the atmospheric pressure to 133 Pa, and the third reactor is

10 to conduct the further polycondensation at a temperature of 270° to 290°C under a pressure of 200 to 13.3 Pa.

The above-mentioned objects of the present invention can be attained by using three single reactors each for a esterification step, a prepolymerization step

15 and a final polymerization step, where single reactor without any external power supply are used for at least one of the esterification step and the prepolymerization step, and a stirring power-requiring reactor is used for the final polymerization step. The first reactor for the

20 esterification step according to the present invention is, for example, a natural circulation type evaporator comprising a vertical cylindrical vessel provided with an inlet and an outlet for a raw material feed such as a mixture an aromatic dicarboxylic acid or its derivative

25 and a glycol and an oligoester or polyester at the bottom of the vessel and further with a vapor pipe for discharging the vapors at the top of the vessel; a heating medium jacket covering the outside of the vessel;

and a shell-and-tube type heat exchanger within the
vessel, the tube outside of the heat exhchanger being
heated by a heating medium, while allowing the raw
material feed to rise through the tubes to unnecessitate
5   the external power. It is desirable for the natural
circulation type evaporator to make the average speed of
the liquid feed flowing downwards by natural convection
between the inner wall of the cylindrical vessel and the
outer wall of the shell of the sheel-and-tube type heat-
10   exchanger smaller than the average speed of the liquid
feed rising through the tubes of the shell-and-tube type
heat exchanger and also to provide an entrance space at
the bottom of the shell-and-tube type heat exchanger so
as to allow the internally circulating raw materials to
15   uniformly enter into the tubes.
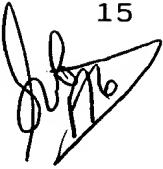
For the second reactor for the
prepolymerization step is used an apparatus comprising,
for example, a substantially vertical cylindrical vessel
provided with an inlet for an oligopolyester or polyester
20   feed from the first reactor and an outlet for a low
molecular weight polyester feed on the lower end side and
at the bottom at the center of the vessel, respectively,
in the longitudinal direction of the vessel, and further
with an outlef for volatile matters at the top of the
25   vessel, and covered with a heating medium jacket on the
outside of the vessel, thereby to unnecessitate any
external power. The second reactor may be further
provided with a heat-exchanging section at the lower part

within the vessel and a residence section with helical
baffle plates at the intermediate part within the vessel
so as to retain the liquid feed and transfer the
polyester feed successively from a lower stage to a upper

5   stage, further with a space for gas-liquid separation at
the upper part within the vessel and a down pipe in the
vertical direction at the center within the vessel so as
to allow the polyester feed to flow downwards as a thin
film.

10   For the third reactor for the final
polymerization step is used an apparatus comprising, for
example, a horizontal cylindrical vessel provided with an
inlet for the low molecular weight polyester feed from
the second reactor at the lower part at one end and an

15 outlet for a high molecular weight polyester at the lower
part at the other end, respectively, of the vessel in the
longitudinal direction thereof, and further with an
outlet for volatile matters at the top of the vessel and
a stirring rotor, which is extended in the longitudinal

20 direction of the vessel and rotates in the vicinity of
the inner periphery of the vessel, the stirring rotor
within the vessel being divided into a plurality of
stirring blocks according to the viscosity level of the
polyester feed, and the stirring rotor being without a

25 rotating shaft but with stirring vanes at the center part
of the strring rotor.

BRIEF DESCRIPTION AT THE DRAWINGS

Fig. 1 is a flow diagram showing an arrangement of reactors for a process for continuously producing polyethylene terephthalate according to one embodiment of the present invention.

Fig. 2 is a cross-sectional view showing an evaporator according to one embodiment of the present invention.

Fig. 3 is a vertical cross-sectional front view showing one embodiment of the present invention.

Fig. 4 is a vertical cross-sectional front view showing one embodiment of the present invention.

Fig. 5 is a cross-sectional view along the line A-A of Fig. 4.

Fig. 6 is a cross-sectinal view along the line B-B of Fig. 4.

Fig. 7 is a cross-sectional view along the line C-C of Fig. 4.

Fig. 8 is a cross-sectional view along the line D-D of Fig. 4.

Fig. 9 is a schematic view showing the flow of a polyester feed in backets in a low viscosity stirring block.

Fig. 10 is a schematic view showing the flow of a liquid feed around the thin hollow disk in a low viscosity stirring block.

Fig. 11 is a schematic view showing the flow of a polyester feed around the hollow disk in an

intermediate viscosity stirring block.

Fig. 12 is a schematic view showing the flow of a polyester feed on the thin hollow disk in an intermediate viscosity stirring block.

5   Fig. 13 is a schematic view showing the flow of a polyester feed in a high viscosity stirring block.

Fig. 14 is a cross-sectional view along the line E-E of Fig. 4.

Fig. 15 is a vertical cross-sectional front
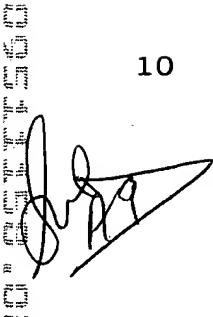10   view showing another embodiment of the present invention.

BEST MODES FOR CORRYING OUT THE INVENTION

Fig. 1 shows one embodiment of the present invention, i.e. a flow diagram showing an arrangement of reactors for a process for continuously producing
15   polyethylene terephthalate. Economically very advantageous industrious process for producing polyester is a direct esterification process, which has been recontly much employed. In Fig. 1, numeral 31 shows a feed mixing tank for mixing and stirring TPA
20   (terephthalic acid) as an aromatic dicarboxylic acid and EG (ethylene glycol) as a glycol in a molar ratio of 1:1.05 to 1:2.0 of the former to the latter both being raw materials for producing polyethylene terephthalate. At this eterification stage of the process, a
25   polymerization catalyst or additives such a stabilizer, a color tone regulator, etc. can be used together, when required.

Polymerization catalyst includes, for example, metal compounds derived from antimony, titanium, germanium, tin, zinc, etc. It is well known that not only the reaction rate, but also hue and heat stability

5 of the resulting polyester depend upon the type and combinations of the catalyst used. Furthermore, the esterification reaction is carried out in the presence of a catalyst at a high temperature for a long time, resulting in occurrence of various side reactions, which

10 discolor the polyester into yellow, increase the content of diethylene glycol (DEG) and the concentration of terminal carboxyl groups beyond the appropriate levels and lower the physical properties such as melting point and strength of polyester.

15 Development of new catalysts has been so far attempted to overcome these problems. Antimony compounds now used most in the industry, particulary antimony trioxide, are distinguished in the cost and effects. However, discoloring of the resulting polyester polymers

20 is inevitable even by using these catalysts. To overcome this disadvantage, a phosphorus-based stabilizer (for example, trimethyl phosphate and triphenyl phosphate) is used together as a stabilizer. In another attempt, an inlet position of the polymerization catalyst or the

25 stabilizer is elaborated to stailize the quality. In the ordinary process, it is preferable to use 200 to 400 ppm of the catalyst and 50 to 200 ppm of the stabilizer.

The raw materials so mixed is led to an

esterification reactor 33 as a raw material feed through

a fed line 32 supplying the raw material feed thereto.

The esterification reactor (first reactor) 33 is provided

with a heating medium jacket structure (not shown in the

5 drawing) on the outer periphery of the reactor to keep

the raw material feed at a reaction temperature and also

provided with a shell-and-tube heat exchange 34 as dipped

in the raw material feed within the first reactor as a

means of heating the raw mateial feed from an external

10 heat source and circulate the raw material feed within

the first reactor by natural circulation, thereby

proceeding with the reaction. Type of most desirably

first reactor is such a calandria type that the

esterification reation is to be carried out by naturally

15 circulating the raw material feed within the reactor on

the basis of utilization of the evaporating action of by-

products formed by the reaction. This type of reactor

needs no external power source for stirring, resulting in

such advantagtes an a simple reactor structure,

20 unnecessity for a stirring shaft-sealing device and a low

reactor-manufacturing cost. One example of such a

desirable first reactor is shown in Fig. 2. Fig. 2 shows

one embodiment of the first reactor. Raw mateial feed 52

flows into a vertical evaporator 51 through an inlet 53

25 provided at the bottom of the evaporator 51 and is heated

while passing through a plurality of heat transfer tubes

in a shell-and-tube type heat exchange 54 to go upwards

by natural convection, where a portion of low boiling

point components in the raw material feed 52 is evaporated, and discharged from a vapor pipe 55 to the system outside. The remaining raw material feed 52 flows downwards between the inner wall of the evaporator 51 and

5 the outer wall of the shell of the shell-and-tube type heat exchange 54 by natural convection and enters into a cylindrical entrance space 56 provided at the bottom of the shell-and-tube type heat exchange 54, where the flow of the raw material feed can be rectified to a less

10 turbulant state, and the average flow speed through the tubes of the shell-and-bube heat exchanger 54 is made higher than the average flow speed of the raw material feed flowing downwards by natural convection between the inner wall of the evaporator and the outer wall of the

15 shell. Thus, the raw material feed can enter into a plurality of heat transfer tubes with a more uniform speed distribution and uniformly heated again to repeat circulation by natural convection. In the course of circulation, the low boiling point components are

20 evaporated and after an appropriate convection time, the resulting oligoester or polyester 59 is led to the system outside from an outlet 60. To generate a smooth, accelarated flow, the flow passage area of the cylindrical entrance space must be designed larger than

25 the total flow passage area of heat transfer tubes and also the passage area of double pipe section formed between the inner wall of the evaporator 51 and the outer wall of the shell of the shell-and-tube type heat

exchanger 54 is made larger than the flow passage area of the entrance space. Numeral 57 shows an inlet of a heating medium and 58 shows an outlet of the heating medium. The outer periphery of the evaporator 51 is

5   covered with a heat insulator or a heating medium jacket (not shown in the drawing). Thus, in the evaporator of this embodiment, uniform flow speed distribution in the axial direction of the heat exchanger can assure uniform evaporation or esterification reaction of the raw

10 material feed, and a better product quality can be effecttively obtained in a short residence time. In case that the raw material feed 52 is a mixture of solid particles and a liquid (which will be hereinafter referred to as slurry), the raw material feed 52, which

15 undergoes natural circulation, flows into the cylindrical entrance space 56 provided at the bottom of the shell- and-tube type heat exchanger 54, and by more smooth rising along a conical member 62, the solid particles are never precipitated at the bottom. That is, when the

20 liquid is a slurry, solid particles contained in the slurry can be prevented from precipitation by providing at the bottom of the evaporator the conical member for rising the raw mateial feed to undergo internal circulation. The conical member may have some curvature.

25 Thus, the evaporator of this embodiment is effective and more suitale for natural circulation of a slurry, and can produce a reliable polyester product of good quality. The present invention is not limited to this type of

evaporator and use of an evaporator with stirring vanes,
is not objectinable for process reasons.

In the first reactor, water foumed by the
reaction is in a vapor form, and forms a gas phase 65

5   together with vaporized EG. Recommendable reaction
conditions for the first reactor are desirably a
temperature from 240° to 285°C and a pressure of the
atmospheric pressure to 3 x 10^6 Pa. The gases in the gas
phase 65 are separated into water and EG by a

10 rectification column (not shown in the drawing), and
water is discharged to the system outside, whereas EG is
returned to the system.

Advantages of the present invention are such
that only a single rectification column is satisfactory,

15 because the esterification step is carried out in a
single reactor, and thus not only the rectification
column manufacturing cost but also number of piping,
valves, control units, etc. can be reduced, resulting in
a large reduction of apparatus cost.

In Fig. 1, the raw material feed, when
retained in the esterification reactor 33 (first reactor)
for a predetermined reaction time reaches a predermined
rate of esterifcation to produce an oligoester or
polyester having an average degree of polymerization of 3

25 to 7 and then the resulting oligoester or polyester is
supplied to an initial polymerization reactor (second
reactor) 37 through a connecting pipe 36, where the
oligoester or polyester feed is heated to a predetermined

reaction temperature by a heat exchanger 38 to undergo
polycondensation reaction, thereby increasing the average
degree of polymerization. The reaction conditions are a
temperature of 250° to 290°C and a pressure of the

5 atmospheric pressure to 133 Pa, preferably 266 Pa to 133
Pa, whereby the polycondensation reaction is made to
proceed to produce a low molecular weight polyester
having an average degree of polymerization of 20 to 40.
The initial polymerization reactor shown in this

10 embodiment is a reactor without stirring vanes, but the
present invention is not limited to such a reactor. The
initial polymerization stage is such that kinetics of
polymerization reaction is rate-determining and thus the
reaction can proceed smoothly only by supplying the

15 necessary quantity of heet for the polycondensation
reaction fully. Thus, there is no need for the
oligoester or polyester feed to unnecessitate the
stirring action by stirrng vanes and it is only necessary
to discharge EG formed by the polycondensation to the

20 system outside. Most suitable second reactor for such an
operation is desirably an apparatus shown in Fig. 3.

In Fig. 3, numeral 71 is a vertically long,
cylindrical vessel whose outer periphery is covered by a
heating medium jacket 72 having an inlet 80 for a heating

25 medium and an outlet 81 for the heating medium, and a
down pipe 73 having an open top end is provided in the
longitudinal direction at the center of the vessel 71. A
plurality of heat transfer tubes 74 are provided at the

lower part within the vessel 71 in parallel to the down

pipe 73. A plurality of helical baffle plates 75 are

provided on given different levels on the outer periphery

of the down pipe 73 above the heat transfer tube 74. The

5 baffle plates 75 each have a clearance 83 between its

edge and the inner wall of the vessel 71 to allow

volatile matters to pass therethrough and partitions the

interior of the vessel 71 in the vertical direction to

form a plurality of residence compartiments 84. There is

10 a space 76 for separating the liquid feed from volatile

matters at the upper part of the vessel 71, that is,

above the down pipe 73 and the uppermost baffle plate

75C. A plurality of tapered liquid receivers 88 are

provided on given different levels on the inner wall of

15 the down pipe 73, through which the oligoester or

polyester feed flows downwards as a thin film. The

oligoester or polyester feed flowing downwards through

the down pipe 73 can be retained once in the individual

liquid receivers and then successively move downwards,

20 and thus the polycondensation reaction can proceed, while

minimizing the oligopolymer or polyester feed short path

and effectively separating the vapors of volatile

matters.
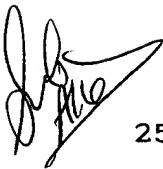
In the second reactor, the oligoester or

25 polyester feed continuously supplied through an inlet

nozzle 77 enters into heat transfer tubes 74 and rises

therethrough while being heated and reaches the lowermost

residence chanber 84A. While the feed slowly rises

through the residence compartment 84A, polycondensation
reaction proceeds and the resulting volatile matters such
as ethylene glycol, etc. move upwards through the
clearances 83 at the outer edge of the baffle plate 75.

5 On the other hand, the feed rises along the helical
configuration of the baffle plate 75 in a whirling stream
in the residence compartment 84A and enters into next
upper residence compartment 84B. Since the feed can move
smoothly in a whirling stream into the next upper

10 residence compartment 84B, the feed can rise successively
through other upper residence compartment without causing
any backflow, and thus the polycondensation reaction
proceeds effectively.
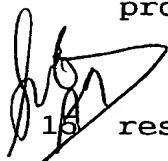
The feed, which has reached the uppermost
15 residence compartment 84C, flows over the top edge 82 of
the down pipe 73 and flows downwards along the inner
periphery of the down pipe 73 as a thin film, while
separating vapors of volatile matters resulting from the
reaction, and thus the polycondensation reaction can

20 further proceed. The feed at an advanced stage of
polycondensation reaction, which has been separated from
vapors of volatile matters resulting from the reaction,
is dischaarged to the system outside through an outlet
nozzle 78, while the resulting volatile matters are

25 separated from entrainments of the feed (polyester) in an
upper space 76 in the vessel 71 and discharged to the
system outside through an outlet nozzle 79 for the
volatile matters.

At that time, the volatile matters are liable to entrain the feed (polyesters), that is, an entrainment problem is liable to occurs. In the present invention, the feed and volatile matters that bump upwards can be

5 shifted towards the circumferential direction by the helical buffle plates 75 to suppress the entrainments. The volatile matters generated in the second reactor, that is, EG, are vaporized in the upper space (gas phase section) 76 kept in a reduced pressure atmosphere and

10 discharged to the system outside after condensation through a condenser (not shown in the drawing) provided at the overhead of the vessel 71.

Advantages of the present invention are such that only a single condenser is satisfactory, because the

15 initial polymerization step is carried out in a single reactor, and thus not only the condenser manufacturing cost but also number of pipings, valves, control devices, etc. can be reduced, resulting in a large reduction of apparatus cost.

In Fig. 1, the feed retained in the initial polymerization reactor (second reactor) 37 for a predetermined reaction time is supplied to a final polymerizer (third reactor) 41 through a connecting pipe 40. In the final polymerizer the polycondensation

25 reaction further proceeds under a good surface renewal action of stirring vanes 42 without a stirring shaft at the center to elevate the degree of polymerization, whereby a high molecular weight polyester having an

average degree of polymerization of 90 to 180 is
produced. Suitable final polymerizer (third reactor) is
shown in Figs. 4 and 15, which have most distingnished
surface renewal performance and power consumption
5   characteristics. Owing to a broad viscosity range of the
liquid feed, the conventional final polymerizer is
divided into two apparatuses to conduct the final
polymerization step, whereas in the present invention the
step can be carried out in a single apparatus, resulting
10 in large reduction in the apparatus cost.

In Fig. 1, numerals 35 and 39 in the first
reactor 33 and the second reactor 37 show a liquid-gas
phases, respectively, and numeral 44 in the third reactor
is a driving means to the stirring rotor. The present
15 final polymerizer will be described be low, referring to
Fig. 4, which shows a vertical cross-sectional front view
of the present apparatus.

In Fig. 4, numeral 1 shows a horizontally long
cylindrical vessel, whose outer periphery is covered with
20 a heating medium jacket (not shown in the drawing), and
at whose both ends in the longitudinal direction are
provided shafts 3a and 3b for rotor support,
respectively. A stirring rotor 4 is fixed to and
supported between the shafts 3a and 3b for the rotor
25 support, and the rotor shaft 3a is connected to a driving
means (not shown in Fig. 4, but shown as 44 in Fig. 1),
and has rotor support members 2a and 2b connected to
connecting support rods 5a, 5b, 5c and 5d, as shown in

Figs. 4 and 5 (the number of connecting support rods
depends on the size of the stirring rotor 4 but is 4 in
this embodiment) at both ends of the stirring rotor 4.
The stirring rotor 4 has a plurality of stirring blocks

5   between the support members (end disks) 2a and 2b. The
support member 2a is a member for a low viscosity
polyester feed and the support member 2b for a high
viscosity polyester feed. The support member 2b is
smaller in the outer diameter than the stirring rotor 4

10   and has scraping vanes 13a and 13b on the vessel end-
focing side to forward the high molecular weight
polyester staying on the inner wall of the vessel 1
fowards the product outlet of the vessel 1 by rotation of
the stirring rotor 4. Its detail structure is shown in

15   Fig. 14, which is a cross-sectional view along the line
E-E of Fig. 4. In the low viscosity zone of the stirring
rotor 4 in the vicinity of an inlet nozzle 11 is provided
a plurality of low viscosity stirring blocks, each of
which comprises a pair of hollow disks 8 with backets

20   formed by scraping vanes 6a and 6b and thin hollow disks
7a provided between a pair of the hollow disks 8 and
being subject to the low molecular weight polyester feed
pouring from the backets (its detailed structure will be
described referring to Figs. 5, 9 and 10). In the

25   intermediate viscosity zone is provided a plurality of
intermediate viscosity stirring blocks, each of which
comprises a pair of hollow disks 8, a plurality of thin
hollow disks 7b having the same outer diameter at equal

distances between the pair of hollow disks 8 and a plurality of scraping vanes 6c provided radially at the outer peripheral side of these disks (its detailed structure will be described, referring to Figs. 6, 7, 11

5 and 12). Furthermore, a high viscosity stirring block, which comprises a plurality of wheel-type disks 9 provided at appropriate distances from one another and scraping vanes 10 provided at the outer peripheral side of the wheel-like disks 9, is provided at the outlet side

10 of the vessel 1 (its detailed structure will be described, referring to Figs. 3 and 13). An outlet nozzle 12 for discharging the high molecular weight polyester product (final polymers) is provided at the lower part at the other end of the vessel 1, and an

15 outlet nozzle 14 for the volatile matters is provided on the upper part of the vessel 1 and is connected to a condenser and a vacuum means through a piping (not shown in the drawings).

In the final polymerizer as shown in Fig. 4, a

20 low viscosity low molecular weight polyester (prepolymer) feed having a low degree of polymerization, e.g. an average degree of polymerization of 20 to 40, continuously fed through the inlet nozzle 11 is at first stirred in a plurality of the low viscosity stirring

25 blocks by backets on the hollow disks shown in Fig. 5. At that time, the feed has such a low viscosity as a few Pas to a few tens Pas. In the low viscosity stirring blocks, one pair of the scraping vanes 6a and 6b forms

one backet at the outer peripheral side of the hollow

disk 8, as shown in Fig. 5, so as to scoop up the feed

into the backet by rotation as shown in Fig. 9. That is,

Figs. 9 and 10 schematically show feed-flowing states.

5   At the bottom of each basket of the scraping vanes 6a and

6b is formed a small hole or clearance δ. With rotation

of the stirring rotor, low viscosity feed 91 is scooped

up by the backet, as shown by numeral 100 in Fig. 9, and

the backets are further inclined downwards by further

10 rotation to allow the feed to start to leave the

inwardly, and at the same time to start to leak out

outwardly little by little from the backet through the

small hole on clearance, as shown by numeral 102 in Fig.

9, to form inward and outward liquid films 101 and 102,

15 respectively, from the backet. The feed 101 flowing

downwards inwardly from the backet pours onto the thin

hollow disk 7a provided in the vicinity of the tip end at

the inward side of the backets, as shown by numeral 103

in Fig. 10, to form liquid films on the surface of each

20 of the thin hollow disk 7a and between the adjacent thin

hollow disks 7a at the same time, thereby producing a

larger evaporation surface area. These actions are

repeated by each rotation of baskets to assure a

sufficient evaporation surface area and a better surface

25 renewal. Satisfactory better performance can be obtained

in the low viscosity stirring blocks even at a low

revolution per minutes such as 0.5 rpm to a few rpm (not

more than 10 rpm), and effective reduction of stirring

power consumption can be obtained. By-products evaporated from the feed pass through center holes 20a of the hollow disks 8 and center holes of thin hollow disks 7a and give off through the outlet nozzle 14.

5   The feed treated in the low viscosity stirring blocks for a predetermined residence time has an increased viscosity such as a few tens Pas, and enters into a plurality of the adjacent intermediate viscosity stirring blocks, whose detailed structure is shown in

10   Figs. 6 and 7. The intermediate viscosity stirring blocks each comprises a pair of hollow disks 8, a plurality of thin hollow disks 7b provided therebetween and scraping vanes 6c provided throughout on the peripheral sides of these hollow disks 8 and 7b.

15   Diameter D1 of center hole 20a of the hollow disk 8 as in Fig. 6 and diameter D3 of a center hole 20a of the thin hollow disk 7b as in Fig. 7 are selected to be optimum, depending on the vapor flow rate of by-products evolved from the feed by the further polycondensation reaction.

20   Furthermore, diameter D2 of small circular holes 20b of the thin hollow disk 7b as shown in Fig. 7 is selected to be optimum, depending on the viscosity of the feed and the vapor flow rate of the by-products evolved from the feed by the polycondensation reaction.
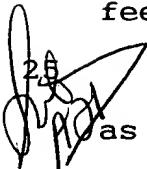
          The feed 92 having an increased viscosity such as a few tens Pas is brought upwards by the scraping vanes 6c by rotation, and flows downwards by grodual inclination of the scraping vanes 6c by further rotation

thereof to form liquid films 104, as shown in Fig. 11.
The liquid films 104 flow down onto the connecting
support rods 5a of the stirring rotor 4 and are retained
there for a long time as sticked thereto suspendewd down

5 therefrom, and furthermore the feed so brought upwards
are suspended downweards over the center hole 20a of the
hollow disk 8 to form a liquid film 105, as shown in fig.
11. A liquid film 107 is likewise formed over the center
hole 20a of the thin hollow disk 7b and also the feed is

10 suspended down over the small circular holes 20b of the
thin hollow disk 7b to form liquid films 106, as shown in
Fig. 12. The feed is formed into such liquid films and
can have much increased degree of polymerization and
viscosity due to much increreased evaporation surface area

15 and surface renewal.
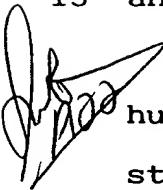
    When the viscosity of the feed reaches a few
hundred Pas, the feed is treated in a high viscosity
stirring block. The high viscosity stirring block
comprises a plurality of wheel type disks 9 having

20 scraping vanes 10 (Fig. 4) at the outer peripheral side
thereof, as shown in Fig. 8. The wheel type disks 9 are
connected to one another at predetermined distances by
the connecting support rods 5a, 5b, 5c and 5d in the
horizontal direction. The scraping vanes 10 comprise

25 front side vanes 10a and back side scraping vanes 10b
alternately arranged on one wheel type disk 9 and on
another adjacent wheel type disk 9, respectively, so that
the total lenghth in the herizontal direction of the

scraping vanes on all the wheel type disks 9 is such that the track of one scraping vane 10a on one wheel bype disk can overlap that of one scraping vane 10b on another wheel type disk, when rotated, to scrape the

5 corresponding entire inner wall surface of the vessel 1.

As shown in Fig. 13, the feed 93 having a viscosity of a few hundred Pas is brought upwards by the scraping vanes 10a by rotation of the stirring rotor 4, and the feed so brought upwards flows downwards by

10 rotation of the whese type disk 9 and suspended to form a liquid film 108, and furthermore a liquid film 109 is formed also over the hollow parts of the wheel type disk 9, thereby creating complicated surface shapes of liquid films. When the viscosity of the liquid feed is further

15 increased to reach a few thousand Pas, the amount of the feed so brought upwards by the scraping vanes 10a is increased. When the revolution per minute of stirring rotor 4 is increased in that state, such a retained feed-circulating phenomenon as to bring upwards the feed

20 retained by the scraping vanes 10a again without leaving the vanes occurs. Thus, the stirring rotor 4 must be rotated at not more than 10 rpm. When the viscosity of the liquid feed becomes higher, the optimum rotation must be made slower. According to the inventors' test, the

25 optimum rotation is in a range of 0.5 to 6 rpm. Stirring and surface renewal actions are repeated, as described above, to promote polycondensation reaction. Votalite matters formed by the reaction pass through the hollow

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parts, center holes and small holes of the hollow disks
successively and move in the longitudinal direction
through the vessel 1 and discharged to the system outside
through the outlet nozzle 14 for the volatile matters.

5   The resulting final high molecular weight polyester
having a high degree of polymerization such as an average
degree of polymerization of 90 to 180 and a high
viscosity such as a few thousand Pas so obtained is
discharged to the system outside through the outlet

10 nozzle 12. At that time the final polyester having such
a high viscosity is liable to stay in the region above
the outlet nozzle 12, but the outer diameter of the
support member 2b of the stirring rotor 4 is smaller than
the outer diameter of the stirring rotor 4, the final

15 polyester never stays on the support member 2b.
Furthermore, the scrapimg vanes 13a and 13b are provided
on the support member 2b on the vessel inner end wall-
facing side, as shown in Figs. 4 and 14 to, forward the
final polyester therefrom towards the outlet 12 of the

20 vellel 1, and thus the inner end wall surface of the
vessel is kept always in a substantially complete self-
cleaned state to prevent the final polyester from the
deposition and staying thereon.
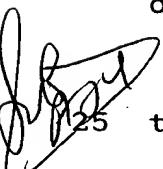
In the polycondensation of polyethylene
25 terephthalate in such a final polymerizer, a prepolymer
feed from the second reactor is continuously supplied
into the final polymerizer (third reactor) through the
inlet nozzle 11, strirred by the stirring rotor 4, while

renewing the prepolymer feed surfaces, and evaporating and removing volatile matters resulting from the polycondensation reaction such as ethylene glycol, etc. and as a result of the polycondensation reaction final

5 polyester of a higher viscosity can be obtained. Volatile matters separated during the polycondensation reaction, for example, ethylene glycol, etc. are discharged to the system outside through the outlet nozzle 14. Operating conditions for the final

10 polymerizer are a temperature of 260° to 300°C, preferably 270° to 290°C, a pressure of 10,000 to 10 Pa, preferably 200 to 13.3 Pa and a slirring rotor revolution of 1 to 10 rpm. The resulting final polymers are discharged to the system outside through the outlet

15 nozzle 12. The interior of the vessel 1 is stirred and kept always in a substantially complete self-cleaned state during the polycondensation reaction and the feed is subject to better surface renewal, and thus final polyester of good quality can be efficiently obtained

20 without any deterioration due to the staying in the vessel 1. This type of the present final polymerizer can be likewise applied to continuous bulk polymerization of resins of polycondensation series such as polyethylene naphthalate, polyamide, polycarbonate, etc.

25   If the viscosity of the feed to be supplied to the final polymerizer is relatively high, no such low viscosity stirring blocks can be used. That is, the low viscosity stirring blocks can be omitted, from the final

polymerizer as shown in Fig. 4. Fig. 15 shows one

embodiment of such a final polymerizer, which is

identical in the structure with that of Fig. 4 except

that the low viscosity stirring blocks are omitted from

5   that of Fig. 4. In Fig. 15, the high viscosity stirring

block comprises a plurality of wheel type disks 9

provided at appropriate distances from one another,

scraping vanes 200 being provided as connected between

the outer peripheral sides of a pair of adjacent wheel

10 type disks 9 but the position of scraping vanes 200 at

one pair of the adjacent wheel type disks 9 is alternated

with that at another pair of the next adjacent wheel type

disks 9.

In the above-mentioned apparatus structure for

15 poducing polyethylene terephthalate, number of reactors

is decreased, as compared with the conventional apparatus

structure and thus the apparatus cost can be much reduced

in the present invention.

Not only number of accessary distillation or

20 rectification columns and condensers can be decreased due

to the decreased number of reactors but also pipings

connecting thereto, instrumentation or control devices

and values can be largely saved. Utility costs relating

to a vacuum means and a heading medium facility can be

25 also largely lowered, resulting in a lower running cost

as an advantage.

According to the present invention, an

apparatus for continuously producing polyester can be

made only from 3 reactors, each directed to the esterification step, the prepolymerization step and the final polymerization step, and can be operated economically due to an increased efficiency throughout

5   the apparatus and energy saving throughout the plant facility.   Total reaction time throughout the first reactor, second reactor and third reactor is 4 to 8 hours.